(12) UK Patent Application (19) GB (11) 2 075 859 A

(21) Application No 8110423
(22) Date of filing 2 Apr 1981
(30) Priority data
(31) 8014039
(32) 29 Apr 1980
(33) United Kingdom (GB)
(43) Application published
25 Nov 1981
(51) INT CL³
B01J 8/02
(52) Domestic classification
B1F D1E
(56) Documents cited
GB 2055606A
GB 1574723
GB 1538082
GB 1387044
GB 1356151
GB 1307845
GB 862410
GB 660540
GB 467147
GB 364215
GB 306883
(58) Field of search
B1F
(71) Applicants
Humphreys and Glasgow Limited,
22 Carlisle Place,
London, SW1P 1JA.
(72) Inventors
Andrew Arthur Deane
(74) Agents
Boult, Wade & Tennant,
27 Furnival Street,
London, EC4A 1PQ.

(54) **Catalytic reactor with internal heat exchanger**

(57) A reactor and associated method for performing exothermic or endothermic gas-phase catalytic reactions uses at least one, usually fixed, bed of catalyst whilst transferring heat to or from another medium by means of a process gas stream flowing from or to the or each bed by way of a tubular heat exchanger within and extending axially through the catalyst bed(s). Preferably the stream is directed via annular passages to a central tubular heat exchanger. The gas side of the exchanger is usually divided into sections by intermediate tube-sheets. The reactor is particularly suitable for methanol synthesis from hydrogen and oxides of carbon; the heat being recovered as steam produced in the central tubular heat exchanger.

GB 2 075 859 A

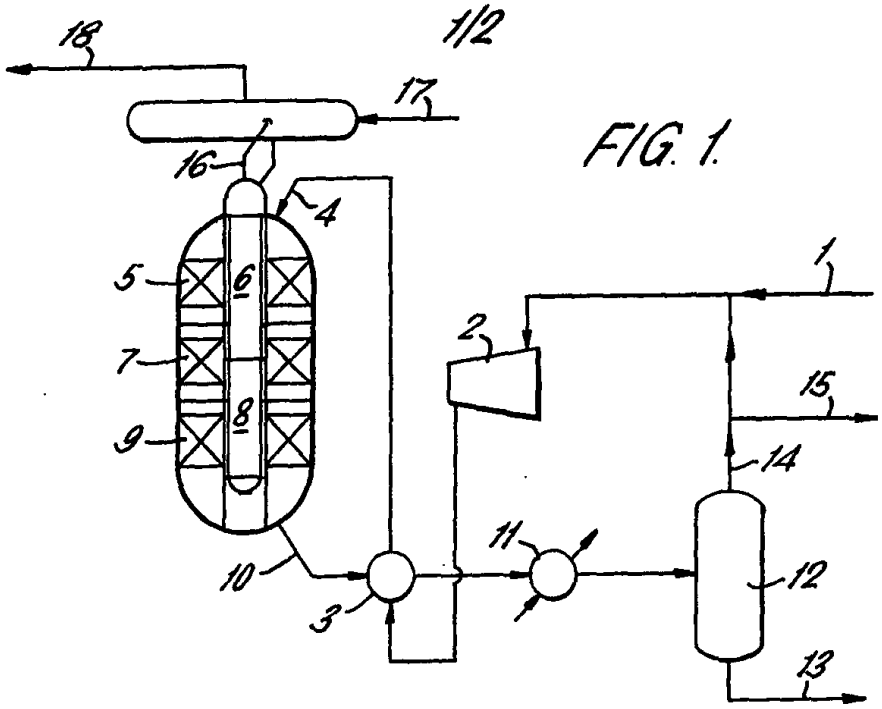

18
17
16
4
5
6
7
8
9
10
3
2
1
15
14
11
12
13

FIG. 1.

FIG. 3.

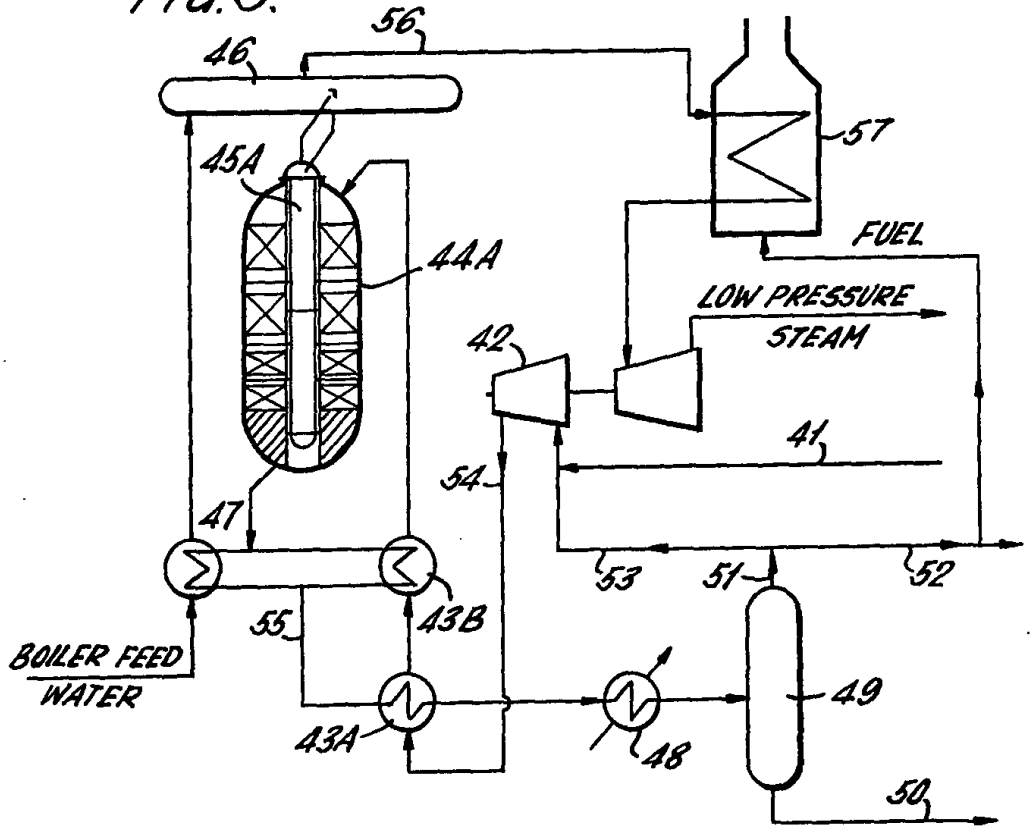
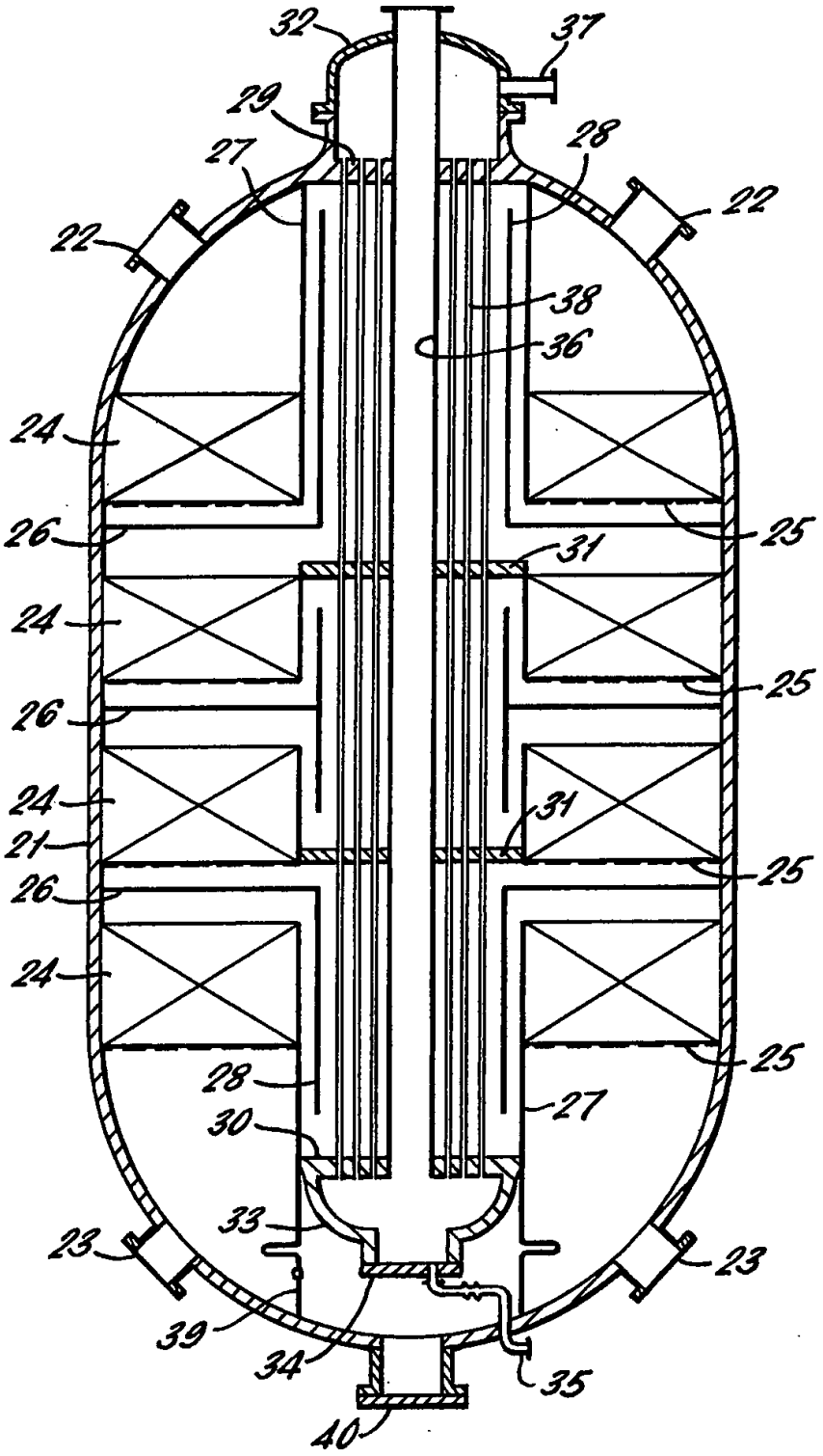

56
46
45A
44A
57
FUEL
LOW PRESSURE
STEAM
42
41
54
47
53
51
52
55
43B
BOILER FEED
WATER
43A
48
49
50

FIG. 2.


32
37
29
28
27
22
22
38
36
24
26
25
31
24
25
26
24
31
21
25
26
24
25
28
27
30
33
23
23
39
34
35
40

SPECIFICATION

Catalytic reactor with internal heat exchanger

This invention relates to a catalytic reactor and associated method for performing gas-phase catalytic reactions with associate heat transfer internally of the reactor.

There are many commercially-operated processes which involve the exothermic chemical reaction of gases at high pressure over solid catalysts. These processes include methanol synthesis from hydrogen and oxides of carbon, ammonia synthesis from hydrogen and nitrogen, methane synthesis from hydrogen and oxides of carbon, conversion of carbon monoxide to hydrogen by the water-gas shift reaction, synthesis of hydrocarbons from methanol, and synthesis of hydrocarbons from hydrogen and oxides of carbon. For all of the above processes conversion is inhibited by the reactants approaching equilibrium with the products unless the natural temperature rise is inhibited by such methods as removal of heat or injection of cold gas. Also, all of the above processes are very energy intensive. With increasing energy costs there is a greater need to recover energy at as high a level as possible.

Reaction vessels for these processes have become increasingly large over the last few years for the following reasons. For traditional high pressure processes, such as ammonia and methanol synthesis, reaction pressures have been reduced in order to save compression energy. Because of the higher increase of energy costs compared with equipment costs more catalyst is being used to get more conversion from the same gas flow. In order to save on compression power, wider vessels have been used to reduce the pressure drop through the catalyst. There has been a move to larger capacity plants and this is likely to be increased as upstream units such as steam reformers, partial combustion units and coal gasification units increase in size.

At present there are several different designs of reactor operating in exothermic catalysis processes. There are also four methods used for controlling the gas temperature rise.

The simplest and least efficient method is to recycle large volumes of inert gas through the catalyst. This does not require any special design of reactor but can be very costly in compression energy, and it is difficult to recover much of the reaction heat because of the need to preheat large volumes of gas.

Quench reactors have been used for many years for ammonia and methanol synthesis. In these, the process gas temperature is controlled by the injection, at intermediate levels, of a part of the feed gas at a lower temperature. Quench reactors have the advantage of being relatively simple and of having very low pressure drops. Their disadvantages are that they are limited on percentage conversion and that heat recovery is external to the reactor and at comparatively low temperature.

For higher conversions, reactors with internal feed gas pre-heaters have been used particularly for ammonia synthesis. Here gas is cooled between its passage through beds of catalyst by preheating the feed gas to the first bed. Although there are fewer limitations on conversion than with a quench reactor, the level of heat recovery is similar and there are considerable problems both in adjusting the dimensions of the heat exchangers to fit the dimensions of the catalyst beds, and in making the heat exchangers and catalyst beds accessible for maintenance purposes.

The fourth method for controlling temperature is to recover reaction heat by transferring it to some secondary fluid within the vessel. The only commercially-successful design of this type, for a fixed-bed catalyst (as opposed to a fluidised catalyst), comprises a tubular reactor with its catalyst disposed in the tubes and boiling water distributed within the cylindrical shell. This reactor was developed for use in the synthesis of hydrocarbons from hydrogen and carbon monoxide and has been used successfully for methanol synthesis. The major advantages of this reactor are the high temperature of heat recovery and potentially high conversions. The major disadvantages of this design are, however, the high pressure drop of the gas through the tubes and very high cost of the vessel. Only about 30% of the vessel volume is available for the catalyst, and the reactor requires two, very large, high pressure tube-sheets. The production capacity of this kind of reactor is therefore severely limited. Furthermore, there is no simple means of access to the tubes without the ingress of air onto the catalyst; all of the catalysts for the above-mentioned processes being intolerant of exposure to air.

Thus the requirement of high temperature heat recovery in a relatively high capacity reactor is not well met by any of the reactors used in present technology. In order to get a high capacity from a single reactor it is necessary to use the majority of the reactor cross-section for catalyst. Methods have been proposed for having several catalyst beds within a single pressure vessel and of removing the gas between beds to an external heat exchanger for the recovery of the reaction heat. Such a reactor would, however, require large spaces between the catalyst beds for collecting and redistributing the gas; it would be liable to have a high pressure drop, and would require an additional pressure shell for the heat exchanger.

Designs have been constructed by Montecatini in which heat is extracted by utilizing high-pressure water exchangers which in turn exchange their heat with lower pressure water to raise steam. It is however desirable to be able to raise steam directly without the need for an intermediate fluid.

The present invention relates to a reactor and associated method for carrying out generally exothermic, gas-phase, catalytic reactions in a pressure vessel containing both at least one bed of catalyst and a tubular heat exchanger. Generally a heat exchange fluid such as water or steam is arranged to flow through the tubes of the heat exchanger, but in certain embodiments of the present invention the process gas, rather than the heat exchange fluid, may be arranged to flow through these tubes if found to be advantageous.

One object of the present invention is to provide a catalytic reactor and associated method achieving relatively high production per unit volume of reactor.

Another object of the present invention is to provide a catalytic reactor and associated method having good design flexibility.

A further object of the present invention is to provide a catalytic reactor and associated method having a high level of heat recovery within a single pressure vessel.

Yet another object of the present invention is to provide a catalytic reactor and associated method wherein access to the heat exchange tubes is relatively simple and can avoid the ingress of air onto the catalyst.

An additional object of the present invention is to provide a catalytic reactor and associated method whose unit capital costs are not inordinately high, and wherein adequate allowance can be made for the effects of thermal expansion.

In accordance with the present invention there is provided a catalytic reactor comprising, in the same vessel, at least one catalyst bed for a process gas and a tubular gas/fluid heat exchanger for the gas, wherein the at least one bed and the heat exchanger are disposed around a common reactor axis; one being arranged around the centrally-placed other, with the heat exchanger extending axially through the at least one bed, wherein gas ducting is provided within the reactor vessel to arrange for the flow of gas through the heat exchanger to be co-current with the overall gas flow through the reactor.

Although from a technical point of view one catalyst bed of any sort could be used, the present reactor preferably comprises a plurality of fixed catalyst beds wherein the heat exchanger is sub-divided into separate sections, and the ducting is arranged to guide the flow of gas alternately between successive catalyst beds and their corresponding heat exchanger sections. Generally the reactor vessel is of substantially cylindrical shape and the ducting comprises at least one cylindrical sleeve providing an annular space between the at least one catalyst bed and the heat exchanger. It is advantageous for the at least one sleeve to be individually connected to the vessel at a single longitudinal cross-section of the vessel so as to allow free longitudinal thermal expansion of the at least one sleeve. The or a sleeve may be extended longitudinally, optionally via a flexible shroud, to seal against the internal wall of the vessel. In these preferred arrangements the heat exchanger is of the "shell-and-tube" type; one or more sleeves functioning as the "shell(s)" thereof.

The present reactor is capable of handling either exothermic or endothermic reactions, but the former are generally preferred. In a particular embodiment the heat exchanger is in the form of a natural-circulation steam boiler having a centrally-aligned down-comer for the boiler water; which down-comer, when the reactor includes a plurality of catalyst beds, supports the gas seals necessary to sub-divide up the heat exchanger into sections.

Whilst the general flow of gas through the reactor and through the heat exchanger is defined, it is possible to vary considerably the relative position of the at least one catalyst bed with respect to the heat exchanger, and the path taken by the gas flowing through the at least one bed. In one embodiment, the gas entry and exit points for the at least catalyst bed are arranged longitudinally through the vessel, so that the flow of gas through the at least one catalyst bed is axial with respect to the central axis, wherein the heat exchanger is disposed in a central cylinder around the central axis, and the at least one catalyst bed surrounds the heat exchanger; the gas ducting being disposed between the heat exchanger and the at least one catalyst bed. In another embodiment the gas entry and exit points for the at least one catalyst bed are arranged radially across the vessel, so that the gas flow through the at least one catalyst bed is radial with respect to the central axis wherein the heat exchanger is disposed in a central cylinder around the central axis, and the at least one catalyst bed surrounds the heat exchanger; the gas ducting being arranged in two sections; one section being between the heat exchanger and the at least one catalyst bed, and the other section being between the lateral wall of the vessel and the at least one catalyst bed.

In accordance with general practice, the heat exchanger preferably includes a plurality of baffles and, optionally, of heat exchanger fins on its gas side to enhance heat exchange therethrough. In addition, a portion of the heat exchanger can be adapted to exchange heat with a medium other than its heat exchange fluid and the said gas.

The present invention also provides a process for catalytically reacting a process gas stream which method comprises passing the stream through a catalytic reactor of the present invention. The present invention further provides a product when formed using either a catalytic reactor or a process of the present invention, or a derivative thereof.

Suitable exothermic processes for which this reactor can be used include the formation of ammonia from hydrogen and nitrogen, ethylene oxide from ethylene and oxygen, methane from hydrogen and oxides of carbon, a hydrocarbon from methanol or from hydrogen and oxides of carbon, or hydrogen from carbon monoxide and steam. Here the preferred method for heat removal is by the natural circulation and boiling of water. Endothermic reactions can also be performed using the present reactor, such as methanol or ammonia cracking in which the heat input is via process steam or some other suitable high temperature fluid.

As noted above, present technology requires that in order to recover reaction heat at a relatively high temperature a tubular reactor is commonly used which has a high pressure drop, a high unit cost and can only use about 30% of the vessel volume and about 45% of the vessel cross-section for catalyst. The present invention seeks to overcome these limitations by having the catalyst in at least one, preferably fixed, bed within the pressure vessel and by making the gas passing through the or each bed flow by a particular route through an internal, tubular heat exchanger. The heat exchanger is usually divided into sections (with a plurality of

catalyst beds, generally one less than the number of beds) by intermediate tube-sheets. The heat transfer fluid preferably flows through the tubes of the heat exchanger. By this method a large volume of gas may be processed in one reactor using about 60% of the vessel volume and about 90% of the vessel cross-section for catalyst, and high pressure steam may be raised from the reaction heat, when the reaction is exothermic.

The reactor of the present invention has the characteristic of being simple to fabricate with very few problems of differential expansion. It is also possible to gain access to the heat exchanger tube plates without allowing air to contact the catalyst, so that the ends of a failed tube may be plugged.

Preferably the gas is conducted between the catalyst and the heat exchanger by means of annular spaces in or between the or each catalyst bed and the heat exchanger. These spaces are generally enclosed by two sets of cylindrical shrouds, the outer shroud in one embodiment being also used as a part support for the catalyst, whilst the inner shrouds are preferably connected to radial seal plates between the catalyst beds.

The present invention may be used for either axial flow or radial flow through the catalyst; the reactor vessel generally possessing substantial rotational symmetry about a central axis. With radial flow, an annular space is also required between the or each catalyst bed and the vessel wall, so that the two annular spaces are used for distribution and collection of the gas into and out of the catalyst, as well as for directing the gas to and from the heat exchanger.

For large reactors, particularly with radial flow through the catalyst, the gas flow need not alternate between catalyst and heat exchanger, but some by-passing of either the catalyst or the heat exchanger may sometimes be desirable. Here, for example, the at least one catalyst bed can be sub-divided up into separate sections; further gas ducting being provided therebetween. One advantage of this arrangement is to achieve either a vessel with reduced diameter or a greater flow of gas through a vessel of given diameter.

Figure 1 is a flowsheet illustrating the use of a catalytic reactor of the present invention in a methanol synthesis loop.

Figure 2 is a schematic representation of the longitudinal cross-section of a catalytic reactor of the present invention. This embodiment comprises four annular catalyst beds surrounding a central heat exchanger with axial flow through the catalyst.

Figure 3 is a flowsheet for a methanol synthesis process using a catalytic reactor of the present invention.

Referring to the drawings, Figure 1 shows the use of a three-bed reactor of the present invention in a methanol synthesis loop, and particularly illustrates the steps involved in the process.

In step 1 high pressure make-up gas between 30 and 110 BAR (3 to 11 MPa) is fed to the synthesis loop. This gas contains hydrogen, carbon monoxide and carbon dioxide as well as some inert gases such as methane and nitrogen.

In step 2 the make-up and the circulating gas are compressed to the synthesis pressure of 30 to 110 BAR (3 to 11 MPa) by the circulating compressor.

In step 3 the feed gas to the reactor is heated to reaction temperature by the hog gases leaving the reactor in a tubular heat exchanger.

In step 4 the hot gases are fed to the top of the reactor and distributed for even flow over the whole cross-section of the first catalyst bed.

In step 5 the gas passes over the first bed of synthesis catalyst; some of the hydrogen and carbon oxides being converted to methanol; and the temperature rises from about 450°F (232°C) to about 550°F (288°C).

In step 6 the gas passes over the first section of the heat exchanger. Here the gas is cooled to about 430°F (221°C). The pressure of the water is likely to be between 15 and 30 BAR (1.5 to 3 MPa).

In step 7 the gas passes over the second catalyst bed where more methanol is synthesised and the temperature rises to about 540°F (282°C).

In step 8 the gas passes over the second section of the heat exchanger in a manner similar to step 6. The gas is again cooled to about 450°F (232°C).

In step 9 the gas passes over the third catalyst bed where more methanol is synthesised and the temperature rises to about 530°F (277°C). In this example only 3 beds of catalyst are used.

In step 10 the hot gas leaves the reactor containing about 5% methanol by volume. The gas is cooled by interchange with the feed gas as mentioned in step 3.

In step 11 the hot gas is further cooled by cooling water or air; most of the product methanol and water being condensed.

In step 12 the liquid and gas phases are separated so that the crude methanol product may be removed from the loop in step 13.

The remaining gas, containing usually less than 1% of methanol vapour, leaves the separator in step 14.

Some of the gas is purged from the loop in step 15, in order to keep down the concentration of inerts in the loop, and the remainder of the gas is compressed back to reaction pressure in the circulating compressor, step 2.

Water for the reactor heat exchanger in step 6 and 8 is fed from the steam drum in step 16 where steam is separated from the water. The water flows through the tubes of the heat exchanger in steps 6 and 8 by natural circulation. In this embodiment, the water flows downwards and the mixture of water and steam flow upwards.

Boiler water is fed to the steam drum in step 17 in order to maintain the water level in the steam drum.

Steam leaves the steam drum in step 18 at about 300 psig (2.1 MPa). This steam may then be used in steam turbines, steam heaters, as process steam, etc.

Figure 2 shows a four-bed axial-flow reactor of the present invention comprising a vertical pressure shell, 21, having a cylindrical central section and hemispherical ends. The hemispherical ends are not mandatory, but are general practice for high pressure vessels of large diameter. Gas inlet nozzles, 22, are located in the top hemispherical head and gas

distributors of standard type are used to reduce gas upflow within the top head. Gas exit nozzles, 23, are located in the bottom hemispherical head, but if gas velocities are high a collection pipe may be used for improving the flow distribution of gas through the bottom catalyst bed.

Within the pressure shell there are fixed annular catalyst beds, 24. This drawing shows four beds, but the present invention relates to any number of beds including one. The catalyst is supported on catalyst grids, 25, which have a free area through which the process gas can flow. These grids are fabricated in sections for ease of assembly and removal. Between the catalyst beds are radial seal plates, 26, which are welded to the pressure shell and to the inner shrouds, 28. The seal plates are stiffened in order to withstand the differential pressure through any one section of heat exchanger. Manways are fitted in the seal plates for access and catalyst loading.

The catalyst beds are bounded on the outside by the pressure vessel wall and on the inside by the outer shroudds, 27. These shrouds are welded to the tube plates and they support the inner edge of the catalyst support grids, 25. There are annular gaps between the outer shrouds, 27, and the inner shrouds, 28, through which process gas flows after leaving the catalyst beds in order to enter the heat exchanger section, and vice versa. This annular gap may be maintained by the use of vertical fins welded to the shrouds. The seal plates and the tube-sheets extend radially across the annular gap, and the lengths of the shrouds are such as to leave gaps between the ends of the shrouds and adjacent tube-sheets through which flow the process gases.

The top tube-sheet of the heat exchanger, 29, is connected to the top head of the pressure vessel, preferably by welding, although a flange connection may be used. The heat exchanger tubes, 38, extend from the top tube-sheet, 29, through the intermediate tube-sheets, 31, to the bottom tube-sheet, 30. There are generally two fewer intermediate tube-sheets, 31, than there are catalyst beds, 24; in this case two intermediate tube-sheets serving four catalyst beds. The heat exchanger tubes are sealed into the intermediate tube-sheets, 31, to prevent gas bypassing the catalyst beds.

The top head of the heat exchanger, 32, contains a water and steam mixture from the tubes; a flange or manhole being provided in this head for access to the top tube-sheet. The bottom head of the heat exchanger, 33, contains water from a steam drum, and is normally surrounded by high pressure process gas. A manway, 34, is similarly provided to gain access to the bottom tube-sheets. A blowdown facility, 35, is provided and incorporates a flexible connection to allow for the differential expansion of the heat exchanger and the pressure vessel.

The heat exchanger has a central water downcomer, 36, which is welded to the tube-sheets and supports the intermediate tube-sheets. By using the downcomer, the flow in the heat exchanger tubes, 38, is vertically upwards and natural circulation is promoted. Within the heat exchanger top head, 32, the downcomer may be divided horizontally with a gasket seal, in order to facilitate opening and closing of the heat exchanger top head.

Water and steam from the heat exchanger leave the top head at nozzle 37, and boiling takes place within the heat exchanger tubes, 38. The tubes are interlaced with rods to reduce vibration and increase heat transfer on the gas side. Alternatively baffles may be used.

In order to gain access to the bottom heat exchanger head without allowing the ingress of air onto the catalyst, a shroud, 39, may be used to surround the bottom heat exchanger head. This shroud may be fitted with an expansion bellows and pressure relieving devices, and access to the bottom heat exchanger head is then gained through the manway, 40.

It will be apparent to a skilled worker in the art that the system of annular spaces described above, particularly the described inner shroud, can function as a shell cooling system for the heat exchanger when relatively high temperature catalytic reactions are employed, for example in the synthesis of ammonia.

Example

The following example, referring to Figure 3, relates to the production of 2500 t/day of methanol from synthesis gas derived from coal. For reasons of ease of transport and site access two 50% capacity catalytic reactors are employed.

The make-up gas, 41, contains hydrogen, carbon monoxide and carbon dioxide, in stoichiometric proportions, plus smaller quantities of non-reacting gases such as methane, argon and nitrogen amounting to 1.1% molar. A total flow of 10500 Kgmol/hr of synthesis gas enters the methanol loop after final trace removal of sulphur compounds at the suction of the loop circulator, 42, where it combines with the circulating loop gas. In the circulator, 42, the pressure is raised from 57 BAR to 63 BAR (5.7 to 6.3 MPa), and the gas 54, is heated by heat exchange with some of the reactor effluent, 55, to a temperature of 240°C in the reactor exchangers 43A and B. The hot gas then splits into two equal streams and enters the two parallel reactors 44A and B, which in this example each comprise four fixed catalyst beds with three sections of internal heat exchanger (boiler), 45A and B. (Only reactor A is shown.) The temperature of the gas rises as it flows through each bed, but is cooled by each boiler section down to about 240°C to raise 24 BAR (2.4 MPa) steam in each boiler. The steam is then separated from the boiler water in the steam drum, 46.

The gas from each lowermost catalyst bed then leaves the reactors 44A and B, at about 270°C containing about 6.1% of methanol. Various heat exchange arrangements are possible, but in this example 90% of the gas goes direct to the reactor exchanger, 43B, to heat the incoming feed gas, and the remaining 10% goes to a heat exchanger, 47, to preheat the boiler feedwater. After these two exchangers the two gas streams recombine, 55, to enter the second part of the interchange, 43A, and the gas is then cooled in the loop cooler, 48. The condensed methanol is separated in the catchpot, 49, and the crude methanol, 50, is let down in

pressure for further treatment. From the uncondensed gas, 51, leaving the top of the catchpot, a small purge stream, 52, is removed to control the level of inerts in the loop. The remaining gas, 53, then mixes with the incoming make-up gas, 41, before entering the circulator, 42, with a methanol content of 0.3% molar.

Each reactor 44A and B for this plant is 7.5 metres long with an inside diameter of 5.66 metres. The loop heat produces 123 te/hr of 24 BAR (2.4 MPa). steam, 56, which is then superheated in a heater, 57, fueled in this example by some of the loop purge gas, 52.

CLAIMS

1. A catalytic reactor comprising, in the same vessel, at least one catalyst bed for a process gas and a tubular gas/fluid heat exchanger for the gas, wherein the at least one bed and the heat exchanger are disposed around a common reactor axis; one being arranged around the centrally-placed other, with the heat exchanger extending axially through the at least one bed, wherein gas ducting is provided within the reactor vessel to arrange for the flow of gas through the heat exchanger to be co-current with the overall gas flow through the reactor.

2. A catalytic reactor as claimed in claim 1 comprising a plurality of fixed catalyst beds wherein the heat exchanger is sub-divided into separate sections, and the ducting is arranged to guide the flow of gas alternately between successive catalyst beds and their corresponding heat exchanger sections.

3. A catalytic reactor as claimed in claim 1 or claim 2 wherein the reactor vessel is of substantially cylindrical shape and the ducting comprises at least one cylindrical sleeve providing an annular space between the at least one catalyst bed and the heat exchanger.

4. A catalytic reactor as claimed in claim 3 wherein the at least one sleeve is individually connected to the vessel at a single longitudinal cross-section of the vessel so as to allow free longitudinal thermal expansion of the at least one sleeve.

5. A catalytic reactor as claimed in claim 3 or claim 4 wherein a sleeve is extended longitudinally, optionally via a flexible shroud, to seal against the internal wall of the vessel.

6. A catalytic reactor as claimed in any one of claims 3 to 5 wherein the heat exchanger is of the "shell-and-tube" type; one or more sleeve functioning as the "shell" thereof.

7. A catalytic reactor as claimed in any one of the preceding claims, when arranged for exothermic reactions in the at least one catalyst bed, wherein the heat exchanger is in the form of a natural-circulation steam boiler.

8. A catalytic reactor as claimed in claim 7 wherein the steam boiler includes a centrally-aligned down-comer for the boiler water; which down-comer, when the reactor includes a plurality of catalyst beds, supports the gas seals necessary to sub-divide up the heat exchanger into sections.

9. A catalytic reactor as claimed in any one of the preceding claims wherein gas entry and exit points for the at least one catalyst bed are arranged longitudinally through the vessel, so that the flow of gas through the at least one catalyst bed is axial with respect to the central axis.

10. A catalytic reactor as claimed in claim 9 wherein the heat exchanger is disposed in a central cylinder around the central axis, and the at least one catalyst bed surrounds the heat exchanger; the gas ducting being disposed between the heat exchanger and the at least one catalyst bed.

11. A catalytic reactor as claimed in any one of claims 1 to 8 wherein gas entry and exit points for the at least one catalyst bed are arranged radially across the vessel, so that the flow of gas through the at least one catalyst bed is radial with respect to the central axis.

12. A catalytic reactor as claimed in claim 11. wherein the heat exchanger is disposed in a central cylinder around the central axis, and the at least one catalyst bed surrounds the heat exchanger; the gas ducting being arranged in two sections; one section being between the heat exchanger and the at least one catalyst bed, and the other section being between the lateral wall of the vessel and the at least one catalyst bed.

13. A catalytic reactor as claimed in any one of the preceding claims wherein the heat exchanger includes a plurality of baffles and, optionally, of heat exchanger fins on its gas side to enhance heat exchange therethrough.

14. A catalytic reactor as claimed in any one of the preceding claims wherein a portion of the heat exchanger is adapted to exchange heat with a medium other than its heat exchange fluid and the said gas.

15. A catalytic reactor as claimed in any one of the preceding claims wherein the at least one catalyst bed is sub-divided up into sections; further gas ducting being provided therebetween.

16. A catalytic reactor substantially as hereinbefore described with reference to and as illustrated in Figure 2 of the accompanying drawings.

17. A process for catalytically reacting a process gas stream which method comprises passing the steam through a catalytic reactor as claimed in any one of the preceding claims.

18. A process as claimed in claim 17 wherein the reaction of the stream in the at least one catalyst bed is exothermic and steam is generated from boiler water in the tubes of the heat exchanger.

19. A process as claimed in claim 18 wherein the reaction is the formation of methanol from synthesis gas.

20. A process as claimed in claim 18 wherein the reaction is the formation of ammonia from hydrogen and nitrogen, ethylene oxide from ethylene and oxygen, methane from hydrogen and oxides of carbon, a hydrocarbon from methanol or from hydrogen and oxides of carbon, or hydrogen from carbon monoxide and steam.

21. A process as claimed in claims 18 to 20 wherein the stream is pre-heated within the vessel by a section of the heat exchanger adapted for this

purpose.

22. A process as claimed in claim 17 wherein the reaction of the stream in the at least one catalyst bed is endothermic and a heating fluid is fed to the tubes of the heat exchanger.

23. A process as claimed in claim 22 wherein the reaction is the catalytic cracking of methanol or ammonia.

24. A process for catalytically reacting a process gas stream substantially as hereinbefore described with reference to either Figure 1 or Figure 3 of the accompanying drawings.

25. A product when formed using either a catalytic reactor or a process as claimed in any one of the preceding claims, or a derivative thereof.

Printed for Her Majesty's Stationery Office by Croydon Printing Company Limited, Croydon, Surrey, 1981.
Published by The Patent Office, 25 Southampton Buildings, London, WC2A 1AY, from which copies may be obtained.